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Experiences in Operating a Full-Scale Granular Activated-Carbon System with On-Site Reactivation

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J. DeMarco,[1] R. Miller,[2] D. Davis,[2] and C. Cole,[2]

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[1]USEPA, Cincinnati, OH 45268
[2]Cincinnati Water Works, Cincinnati, OH 45228

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16. ABSTRACT

Granular activated-carbon systems were studied in two modes. In the first mode, the systems were acting as a combined filter and adsorber and the filter beds are referred to as a sand replacement system. These carbon beds received coagulated and settled water and the activated carbon acted both as a filter for carryover solids and as an adsorbent material. The second mode employed carbon for adsorption-only by using conventional sand filters prior to the adsorption process-and is referred to as a contactor system. Major organic contaminants were determined. The emphasis for data gathering in this study focused on the economic evaluation of the use of the granular activated-carbon systems investigated. The collected data provide an estimate of the actual costs involved in design, construction, and operation of a system that could be full scale at some utilities.

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Experiences in Operating a Full-Scale Granular Activated-Carbon System with On-Site Reactivation

J. DeMARCO[1]

U.S. Environmental Protection Agency, Cincinnati, OH 45268

RICHARD MILLER, DENNIS DAVIS, and CARL COLE

Cincinnati Water Works, Cincinnati, OH 45228

Granular activated-carbon systems were studied in two modes. In the first mode, the systems were acting as a combined filter and adsorber and the filter beds are referred to as a sand replacement system. These carbon beds received coagulated and settled water and the activated carbon acted both as a filter for carryover solids and as an adsorbent material. The second mode employed carbon for adsorption-only by using conventional sand filters prior to the adsorption process and is referred to as a contactor system. Major organic contaminants were determined. The emphasis for data gathering in this study focused on the economic evaluation of the use of the granular activated-carbon systems investigated. The collected data provide an estimate of the actual costs involved in design, construction, and operation of a system that could be full scale at some utilities.

A GROWING NUMBER OF ORGANIC POLLUTANTS in drinking water supplies are being identified as potentially harmful to health, and removing these substances from drinking water appears to be desirable (*1*,*2*). A granular activated-carbon (GAC) adsorption unit process is especially applicable for use at locations where drinking water sources are subject to a wide variety and varying load of pollutants. The Cincinnati Water Works

[1]Current address: Cincinnati Water Works, Cincinnati, OH 45228

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(CWW) site was considered to be representative of a type of system that might require GAC treatment. The water intake is located on the Ohio River below six major tributaries, one of which is the Kanawha River. Numerous large chemical plants that contribute to the vulnerability of the source water to contamination are located in the Kanawha River Valley. As additional evidence of this vulnerability, 39 organic spills have been recorded by the utility laboratory during the past 3 years. Accordingly, the Cincinnati Water Works and the United States Environmental Protection Agency (U.S.EPA) are cooperating in a joint research project to assess the performance of a large scale GAC adsorption system.

The principal research objectives of interest to utility personnel are the determination of the ability of GAC to improve their water quality, the best type of carbon to use, the most effective way to use the carbon, and always the total cost to achieve the improved water quality. No single study at one location can provide a definitive answer that will apply to all locations because results from the use of GAC are highly site specific. However, at any given location, the answer to the question of improved quality and most effective use must involve the application of chemical analyses of the water treated and data interpretation to determine the use of GAC that maximizes water quality improvements while minimizing costs.

The research project includes the study of GAC systems in two modes. In the first mode, the systems are acting as a combined filter and adsorber and the filter beds are referred to as a sand replacement system. These GAC beds receive coagulated and settled water and the activated carbon acts both as a filter for carryover solids from the preceding processes and as an adsorbent material for dissolved organics. The second mode of GAC use under investigation employs carbon for the purpose of adsorption-only by using conventional sand filters prior to the adsorption process. The latter carbon mode is referred to as post-filtration adsorption and the carbon units are referred to as a contactor system.

Major Parameters Used in Carbon Evaluation

The sample handling and analytic methods for all organic parameters used for this paper are described in detail by DeMarco et al. (3). Grab samples were used because they serve the dual purpose of determining the actual ambient influent and effluent concentrations for the organic constituents monitored as well as for estimating the organic loading into and out of GAC beds. Weekly samples were used during the phase of study being reported because the more frequent sampling conducted in earlier phases indicated no significant data loss would occur for the overall objectives sought. From the many organic parameters used in the study,

the following were selected for the subsequent discussions of water quality: total organic carbon (TOC), trihalomethane formation potential (THMFP), instantaneous trihalomethanes (InstTHM), simulated distribution trihalomethanes (Sim. Dist. THM), chloroform, and bromodichloromethane. The trihalomethane formation potential was determined by using a terminal trihalomethane (TermTHM) sample that was stored for 7 days at a constant temperature of 29.4°C with a pH buffer of 9.5 and an appropriate InstTHM sample (4). These conditions were selected as the maximum that would be present in the Cincinnati distribution system. InstTHM are the trihalomethanes present in the water at the time of sampling. The Sim. Dist. THM samples were based on TermTHM samples stored for 3 days at ambient distribution system temperature and pH conditions. The Sim. Dist. THM samples are fairly accurate estimates of the THM concentrations that would be present in the distribution system at about 3 days water travel time from the treatment system sampled (3). Considerably more organic analyses were performed, but the concentration and frequency of occurrence of the above mentioned substances provide an adequate example of the evaluation procedures undertaken during this study.

Facilities Description

The study is being conducted at the California Plant of the CWW in Cincinnati, Ohio. The normal water treatment process at the 10.3-m^3/s [235 million gallons/day (MGD)] California plant consists of four sequential unit processes that include long-term presettling followed by coagulation, sedimentation, and sand filtration (Figure 1). The CWW system results in low turbidity water being delivered to the sand filters. Also, the storage provided by the long-term presettling basin is very useful during known spill conditions that frequently occur on river systems. Facilities are available to apply chlorine, aluminum sulfate, and powdered activated carbon to the raw water prior to presettling. The plant also has facilities for feeding chlorine, ferric sulfate, lime, soda ash, and powdered activated carbon between the presettling basins and flocculations basins. Also, chlorine can be added between the sedimentation basin and sand filters as well as between the sand filters and clear wells.

Figure 2 is a schematic of the carbon units under investigation at CWW. Two separate 0.22-m^3/s (5-MGD) sand filters have been converted to sand replacement carbon systems (carbon filters). This conversion was effected by removing the 0.76 m (30 in.) of filter sand and replacing it with 0.76 m (30 in.) of 12 × 40-mesh carbon. A 0.16-m^3/s (4-MDG) carbon contractor (post filtration adsorption system) that treats sand filtered water has also been constructed. These two operating

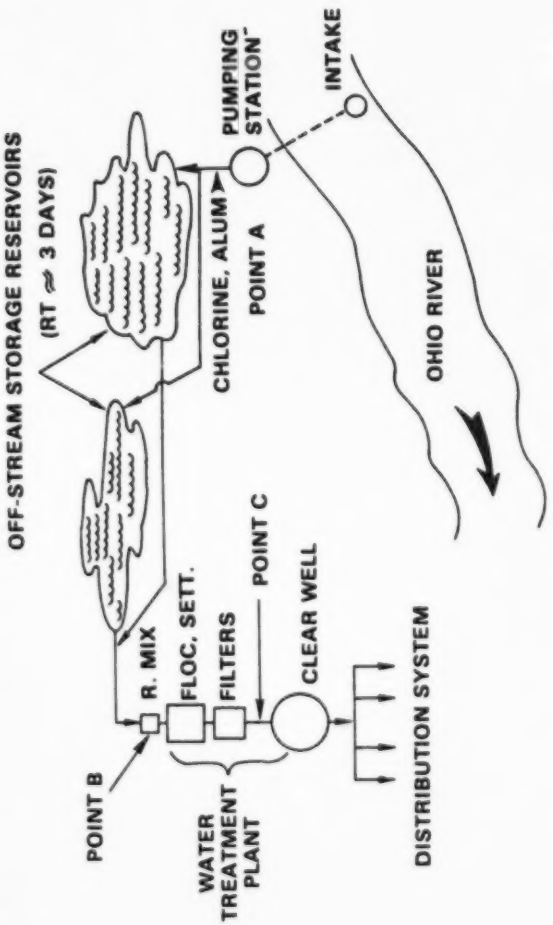

Figure 1. Schematic of Cincinnati Water Works.

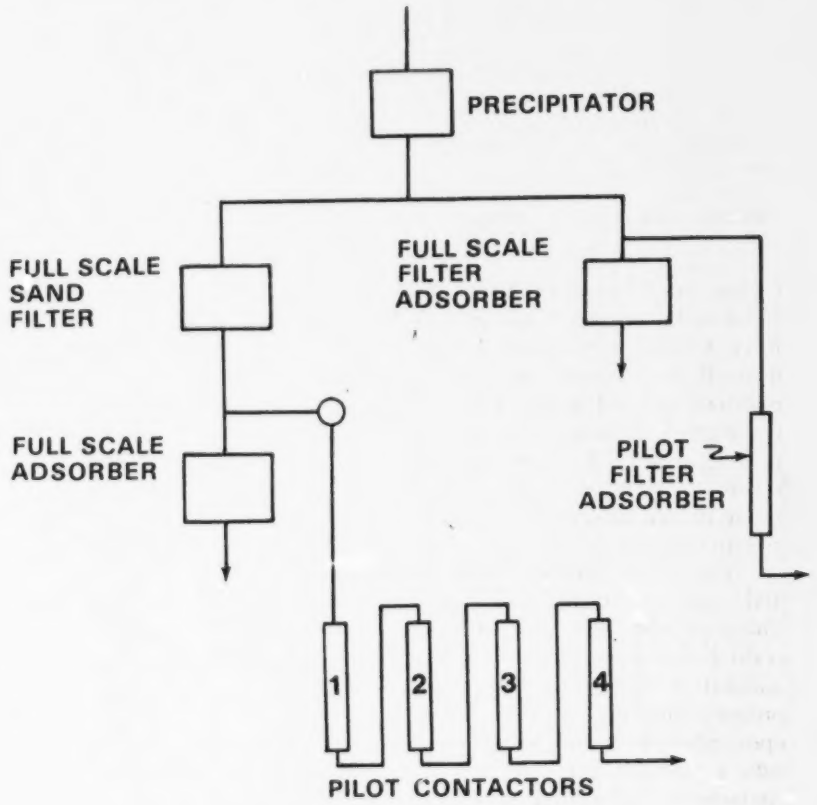

Figure 2. Schematic of carbon units.

modes of carbon application provide a means of comparing the operation of a sand replacement carbon system with that of a post-filtration carbon contactor system. Table I lists pertinent physical data that describe the carbon system now operating. Each of the four 0.4-m^3/s (1-MGD) carbon contactors has sample ports to allow comparison with the sand replacement system at the same empty bed contact times (EBCT). Pilot carbon columns also are being operated parallel to the full-scale systems to assess their ability to simulate the large systems.

Figure 3 is a schematic diagram of the 227-kg/h (526-lb/h) fluid bed carbon reactivation system being used on-site. To date about 455 metric tons (580 tons) of carbon have been reactivated at the utility. Carbon moved from the carbon adsorption system to a 29-m^3 (1050-

Table I. Physical Data for Carbon Systems

Carbon System	*Flow*	*EBCT (min)*	*Carbon Depth (ft)*	*Carbon Weight (lb)*	*Hydraulic Loading (GPM/sq ft)*
Post-filtration contactor	1 MGD	16	14.8	42,500	6.94
Sand replacement	5 MGD	7.5	2.5	96,200	2.5
Pilot post-filtration contactor	920 GPD	16	15.1	34.8	7
Pilot sand replacement	170 GPD	7.8	2.6	3.1	2.5

carbon storage tank. Carbon is transported from the contactors to the spent tank by hydraulically pressurizing the units to provide the motive force. Carbon is transported from the sand replacement systems by a manually fed eductor system. The carbon is then dewatered to about 50% moisture and fed to the fluid bed reactivation furnace. The carbon is reactivated to near virgin properties under a controlled combustion process, cooled by water-quenching, and then placed in a 29-m^3 regenerated carbon storage tank for subsequent transfer to the appropriate carbon adsorption system. Figure 4 is a schematic of the furnace system showing the flow process and air test sample locations.

The carbon contactor shells are constructed of 0.95-cm (3/8 in.) steel and were originally painted inside with 20-mil thick coat of Cook Phenicon 980 with a paint performance specification that requires resistance to carbon abrasion and organic leaching. Because of improper application, the organic leaching requirement had to be enforced and proper refinishing was performed on-site for each contactor with the same epoxy-phenolic paint. Contactor influent and effluent piping is steel pipe with a concrete lining. The metal for sampling probes in the carbon contactor and all wetted, carbon exposed surfaces in the carbon reactivation system battery limits is 316L stainless steel. The 316L stainless steel components in the reactor battery limits have not presented any problems to date. Subsequent leaching tests for the painted interior of the contactors showed no discernible organic contaminants. The contactors have been in operation for about 18 months and parts of the interior surfaces of the contactors appear to require repainting. Additional contactor problems encountered include:

1. Inadequate water to carbon ratio. This was resolved by CWW personnel by adding a water injection point at the base of each contactor.
2. Plugging of the fine mesh screens on the back wash effluent manifold and sampling ports. The problem was resolved by removing these screens with no adverse effects observed to date.

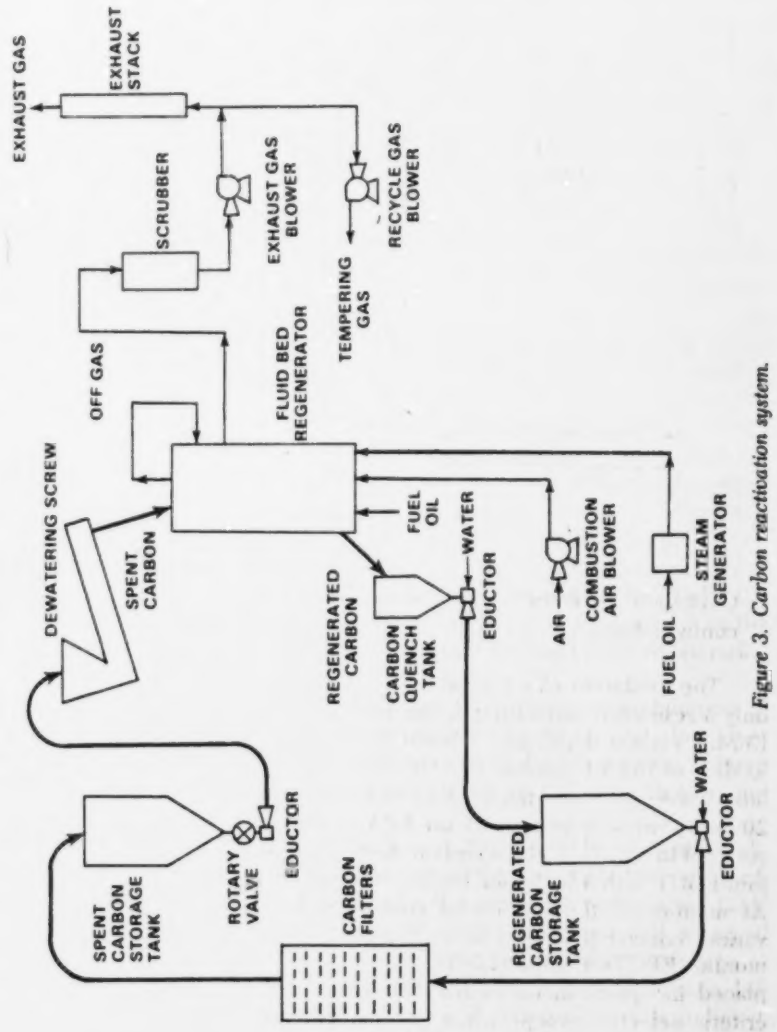

Figure 3. Carbon reactivation system.

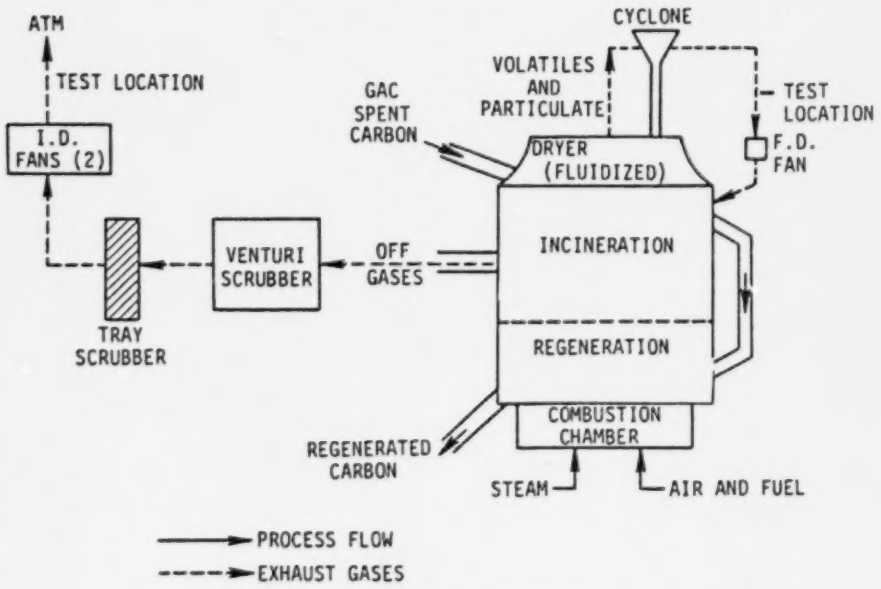

Figure 4. Furnace process flow.

3. Cavitation problems and loss of operating time caused by the piping configuration.

The limitation of a typical sand replacement carbon system is that only a restricted carbon depth can be used. This resulted in the 0.76-m (30-in.) carbon depth and 7.5-min EBCT at the conventional hydraulic loading of 101.8 L/min/m^2 (2.5 GPM/ft^2). Since the post-filtration carbon adsorption system did not have a restriction, an EBCT of between 10 and 20 min was selected based on EPA research data. The final design resulted in a system with a carbon depth of about 4.6 m (15 ft) and a 16-min EBCT with a hydraulic loading of about 285 L/min/m^2 (7 GPM/ft^2). As mentioned, the four metal contactors have multiple ports so that various contact times can be evaluated. Normally, samples are taken to monitor EBCTs of about 4.5, 7.5, and 16.0 min. These two systems were placed in operation on January 14, 1980 and operated following the criteria selected except when prevented by the reactivation system's availability. The two systems were treated like normal water production facilities to the maximum extent possible.

The research project has attempted to maximize data collection for

multiple purposes and the operational criteria for carbon reactivation were: equilibrium between the influent and effluent total THM concentration, and presence of an apparent TOC and/or THMFP steady-state or plateau where the effluent concentration does not equal or exceed the influent concentration but stays approximately parallel to the influent level. The degree of effort expended, for some data collection activities, was greater than what normally would be conducted in a routine water plant operation. Whenever possible, activities that generated atypical costs were excluded from consideration as actual normal costs required to operate the systems.

Results and Discussion

Carbon Adsorption. When beginning to interpret carbon performance, a decision was made on the various types of organic analyses to be used to judge the performance of GAC. When a drinking water concentration of a specific organic substance is required to be controlled within given limits, there exists a definitive use of analytic methods to assess performance. However, Drinking Water Regulations currently contain maximum contaminant levels (MCLs) for only a few organics (*5*, *6*). These are the pesticides endrin, lindrane, methoxychlor, toxaphene, 2,4-D, and 2,4,5-TP, plus total trihalomethanes (chloroform, bromodichloromethane, dibromochloromethane, and bromoform). Obviously all of the general and specific organic analyses currently available could be used in assessing treatment performance. For this study, the organics that were most consistently present and that occurred in the highest concentrations were used to compare the two modes of operation under investigation.

The organics discussed in this paper represent a combination of general and specific compound analyses that attempt to show the extent to which GAC reduces the organic content of drinking water. A more comprehensive organics review was previously provided by DeMarco et al. (*3*).

Comparison of Equal Contact Times. Figures 5–10 present breakthrough curves that show an overview of the concentrations of each measured parameter that passes into the drinking water after GAC treatment by the sand replacement and contactor systems with an equal EBCT of 7.5 min. As expected, the data indicate that a variety of adsorption performances exist, depending on the organic substance, but that in general the organic content of the drinking water was reduced. These figures provide results that satisfy the major objective of comparing the effluent concentrations between the carbon contactor and sand replacement system when operated with the same empty bed contact

times of 7.5 min. Although the flow and hydraulic loadings differed as shown in Table I, the contactors were designed so they could be monitored for various EBCTs, thus the equivalent of the sand replacement EBCT was monitored in the contactors. Figures 5–10 show that very few differences exist in the effluent concentration of the sane replacement and the contactor system. Both systems operated at 192 bed volumes/day. Perhaps a slightly better effluent quality could be claimed for the contactor, based on concentration of organic substance leaving the system. Choosing arbitrary criteria as shown in Table II results in the carbon evaluation data shown. The average influent TOC for both systems was about 1900 μg/L. The average chloroform influent concentration was about 14 μg/L for the sand replacement system and 20 μg/L for the contactor system. The difference in the averages for chloroform is primarily because of the longer contact time with free chlorine residual for the influent to the contactor system. Table II data show that, with equal influent TOC concentrations, the specific criteria of maintaining an effluent concentration ≤ 1000 μg/L resulted in almost the same service time, carbon use rate, cumulative mass loading, and percent removal. Using a percent TOC removal criteria resulted in less uniform comparisons, but no consistent major differences were found that refute the similar adsorptive behavior of the matched EBCT systems. Cumulative mass loading and percent removal data plots also could be used to further prove the similarity of adsorptive peformance for the compared system.

Effect of Various Contact Times. The improved effluent water quality with increasing EBCT is shown by the breakthrough curves of Figures 11–15. Selecting an arbitrary effluent criteria of 5 μg/L for chloroform permits further evaluation of the breakthrough curve data. Such an effluent requirement would result in an operating service time of about 15, 57, and 147 days for EBCTs of 4.5, 7.5, and 16.0 min, respectively, in the contactor system for the time period and conditions shown in Figure 14. Thus, increasing the EBCT by a factor of about 3.6 from 4.5 for 16.0 min increases the operating service time by a factor of almost 10. Also, calculation of carbon use rates for the same criteria shows that they decrease from 572 to 367 to 304 lb of carbon per million gallons of water treated from 4.5 to 7.5 to 16.0 min. The best adsorption carbon use rate for 5 μg/L during the test period apparently would occur at an EBCT longer than 7.5 min and appears to be close to 20 min. However, the total least cost system that meets the design criteria is the one sought.

General Adsorption Evaluation Criteria. The criteria used to assess carbon operation will have an impact on the resulting decisions and must be considered prior to conducting batch, pilot column, or full-scale evaluations. Table III lists additional physical and chemical analyses

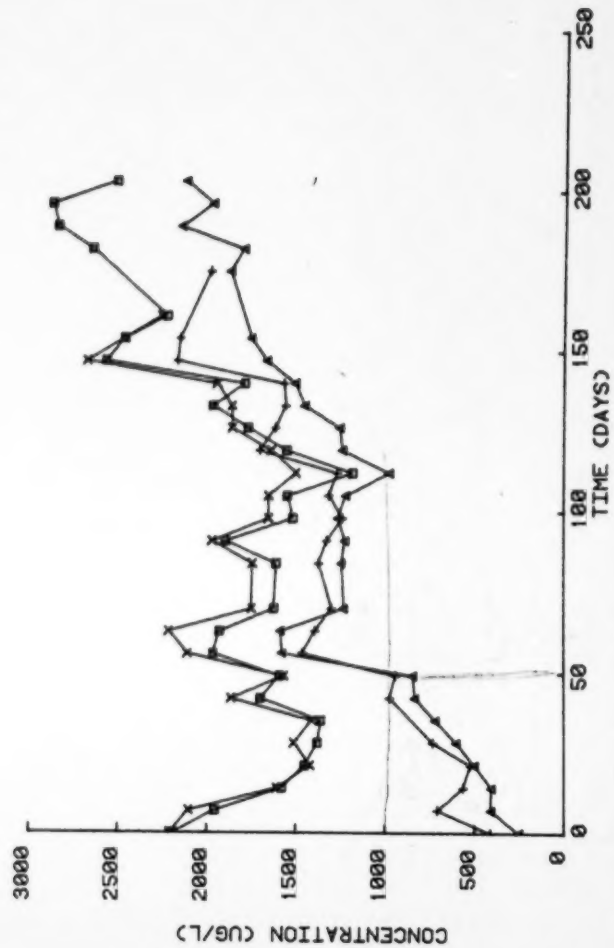

Figure 5. TOC breakthrough curves. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT); × coagulated and settled effluent; and +, sand replacement effluent.

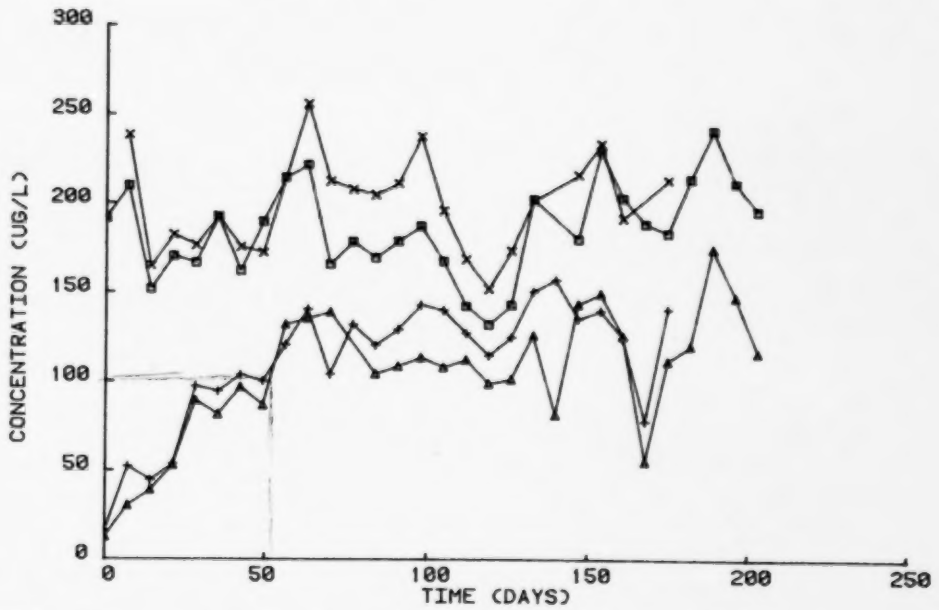

Figure 6. THMFP breakthrough curves. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT); ×, coagulated and settled effluent; and +, sand replacement effluent.

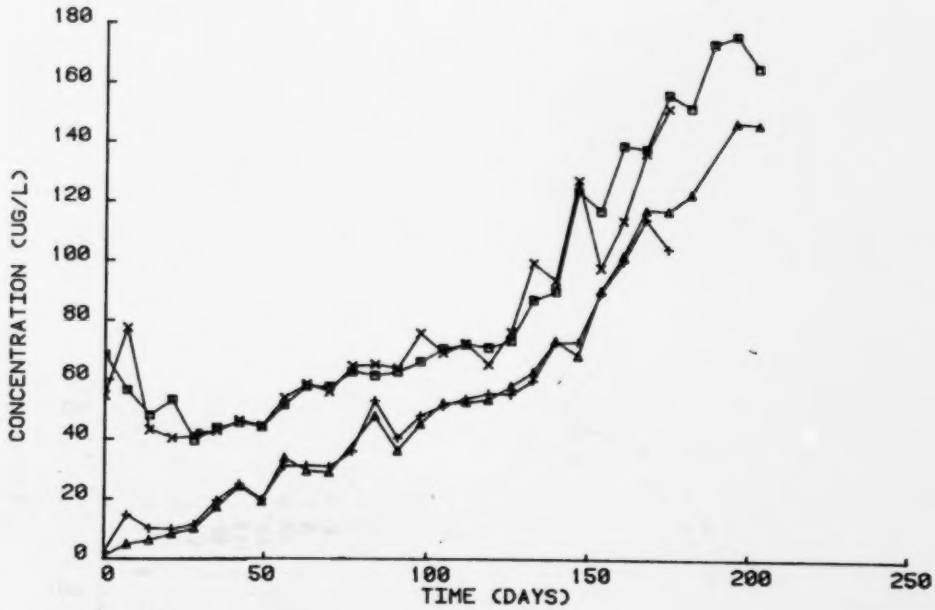

Figure 7. Sim. Dist. THM breakthrough curves. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT), ×, coagulated and settled effluent; and +, sand replacement effluent.

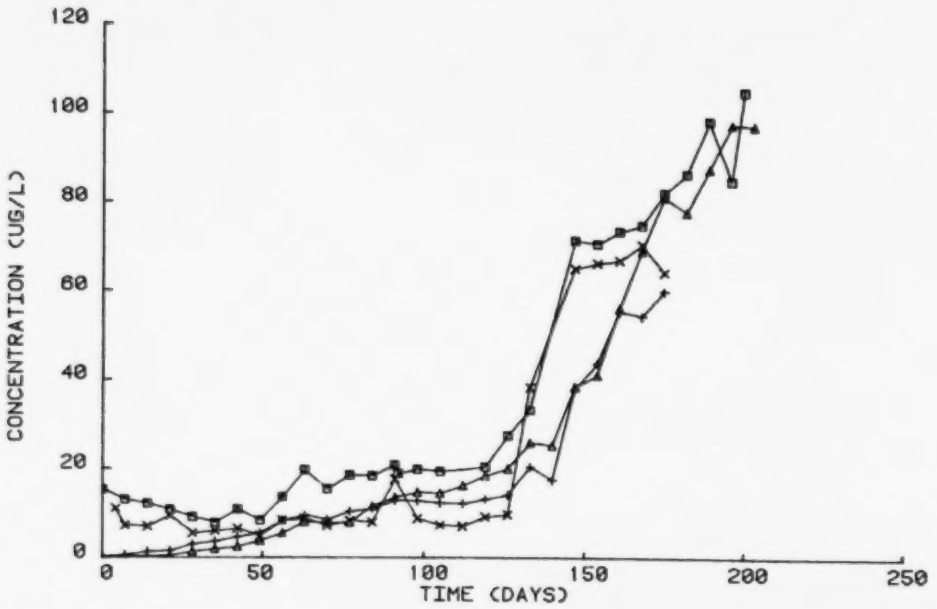

Figure 8. InstTIiM breakthrough curves. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT); × coagulated and settled effluent; and + sand replacement effluent.

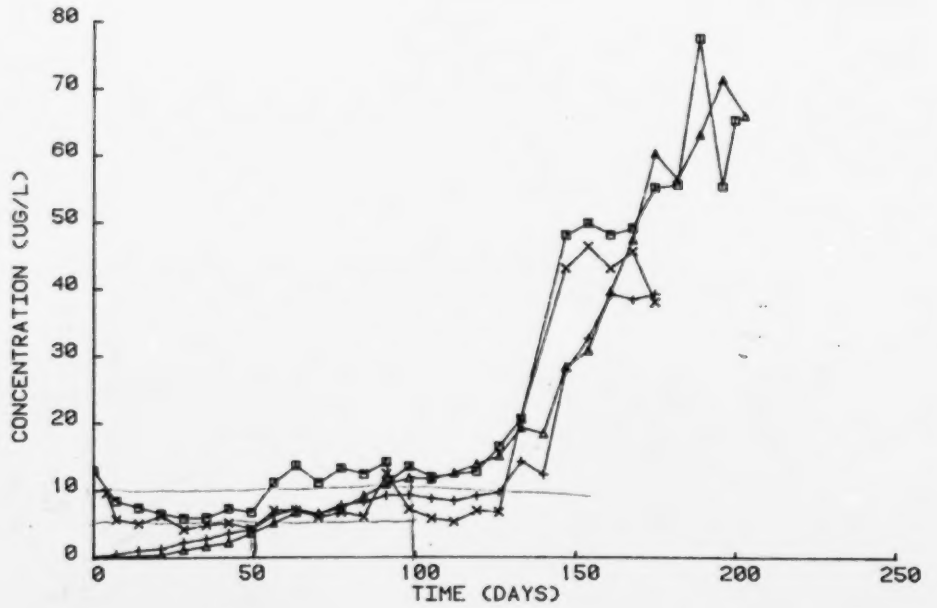

Figure 9. Chloroform breakthrough curves. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT); ×, coagulated and settled effluent; and +, sand replacement effluent.

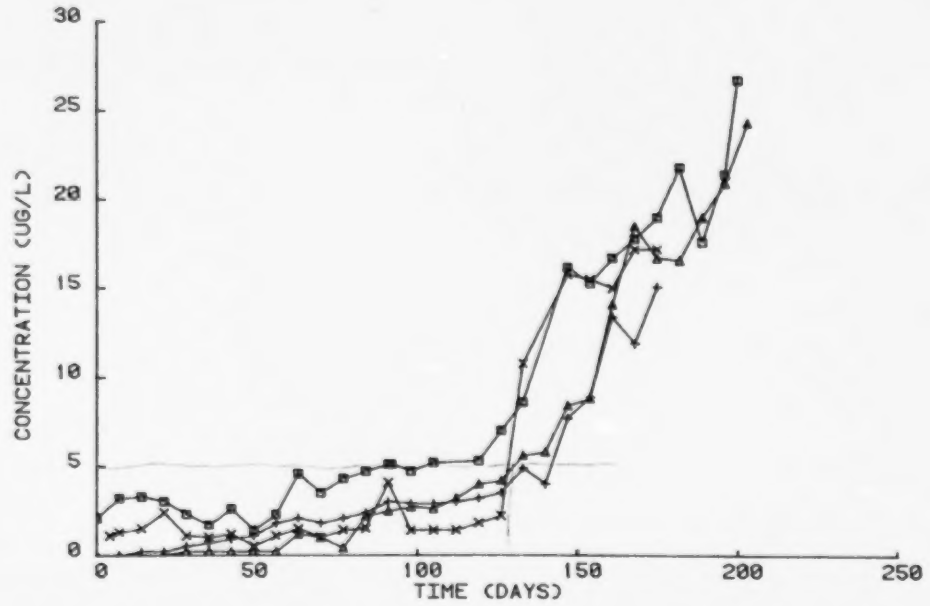

Figure 10. Bromodichloromethane breakthrough curve. Key: □, sand filter effluent; △, contactor effluent (7.5-min EBCT); ×, coagulated and settled effluent; and +, sand replacement effluent.

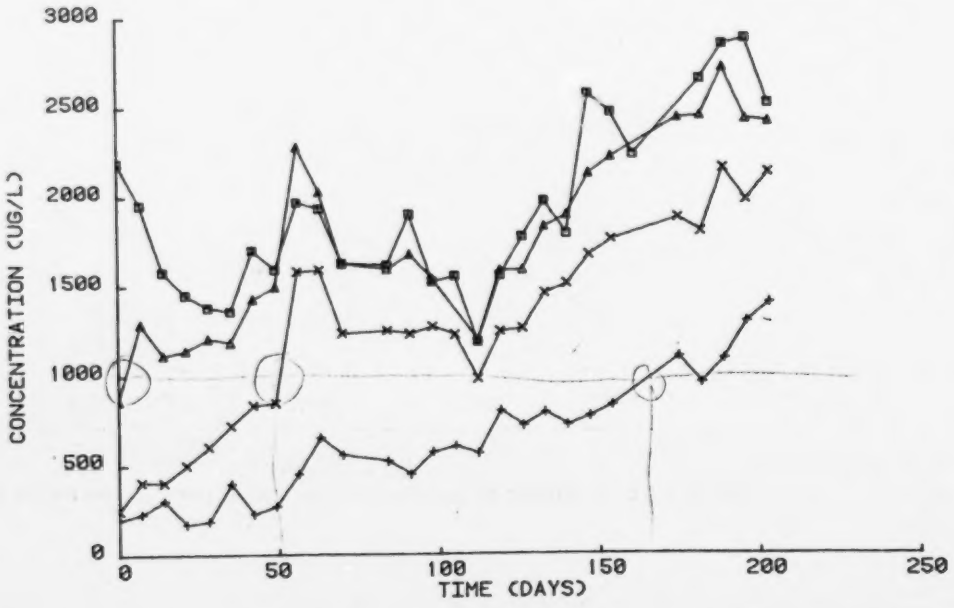

Figure 11. Effect of contact time on TOC. Key: □, sand filter effluent; △, contactor effluent (4.5-min EBCT); ×, contactor effluent (7.5-min EBCT); and +, contactor effluent (16.0-min EBCT).

Table II. Comparison of Sand Replacement (SR) and Contactor (Cont) Carbon Systems Using Various Criteria

	TOC						*Chloroform*					
Evaluation	≤ *1000 μg/L*		≤ *50%*		≤ *75%*		≤ *5μg/L*		≤ *50%*		≤ *75%*	
Criteria	*SR*	*Cont*	*SR*	*Cont*	*SR*	*Cont*	*SR*	*Cont*	*SR*	*Cont*	*SR*	*Cont*
Service time (days)	51	52	37	43	4	8	53	57	29	50	22	36
Carbon use ($lb/10^6$ gallon)	385	405	554	488	4822	2728	361	367	662	420	867	585
Mass load (g/kg)	21.5	21.2	16.9	18.1	2.5	4.9	0.05	0.13	0.037	0.120	0.027	0.096
Percent	40	39	—	—	—	—	10	54	—	—	—	—

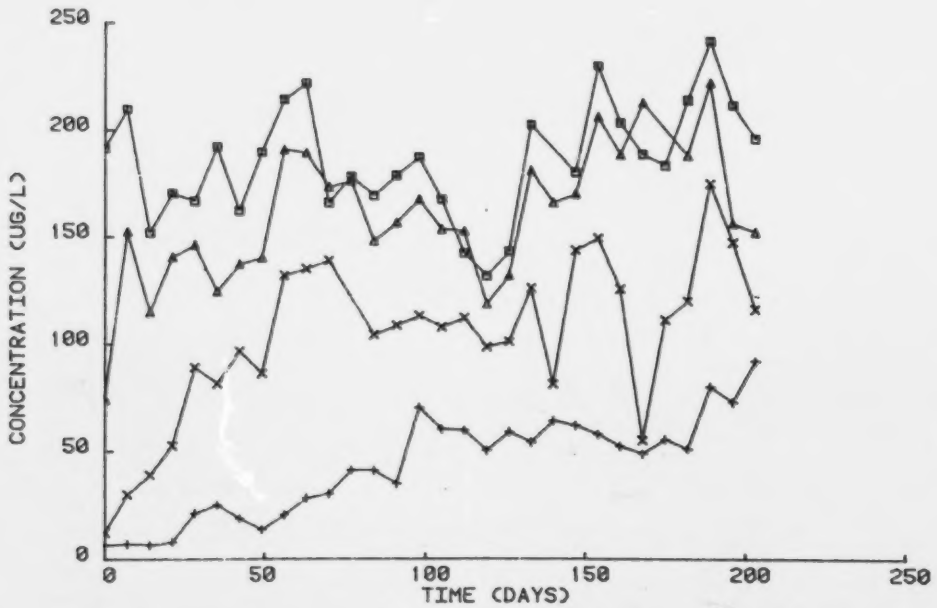

Figure 12. Effect of contact time on THMFP. Key: □, sand filter effluent; △, contactor effluent (4.5-min EBCT); ×, contactor effluent (7.5-min EBCT); and +, contactor effluent (16.0-min EBCT).

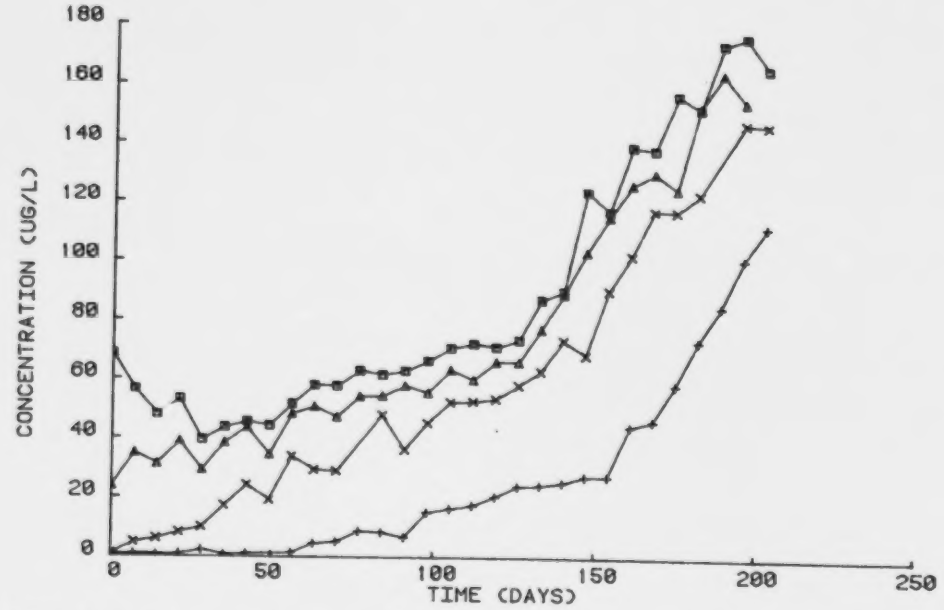

Figure 13. Effect of contact time on Sim. Dist. THM. Key: □, sand filter effluent; △, contactor effluent (4.5-min EBCT); ×, contactor effluent (7.5-min EBCT); and +, contactor effluent (16.0-min EBCT).

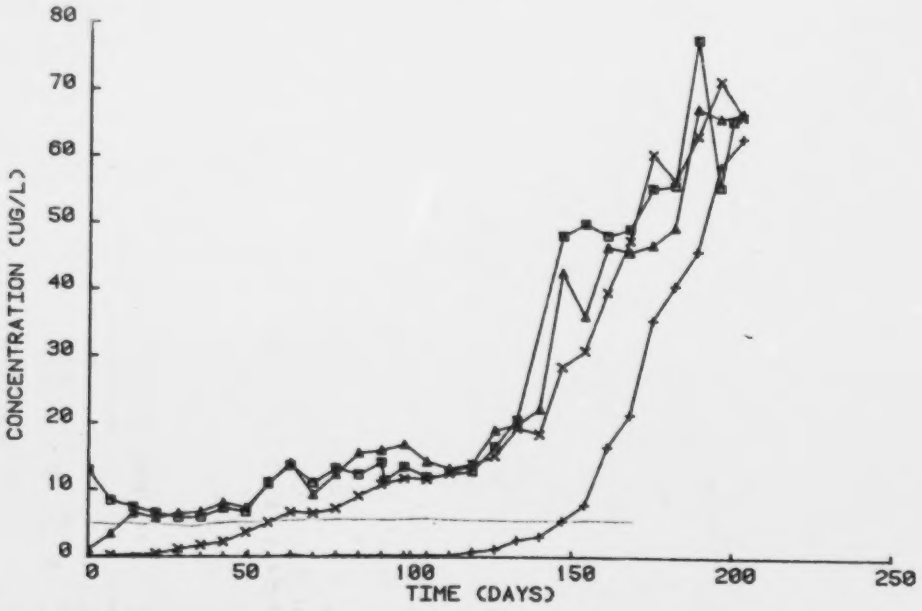

Figure 14. Effect of contact time on chloroform. Key: □, sand filter effluent; △, contactor effluent (4.5-min EBCT); ×, contactor effluent (7.5-min EBCT); and +, contactor effluent (16.0-min EBCT).

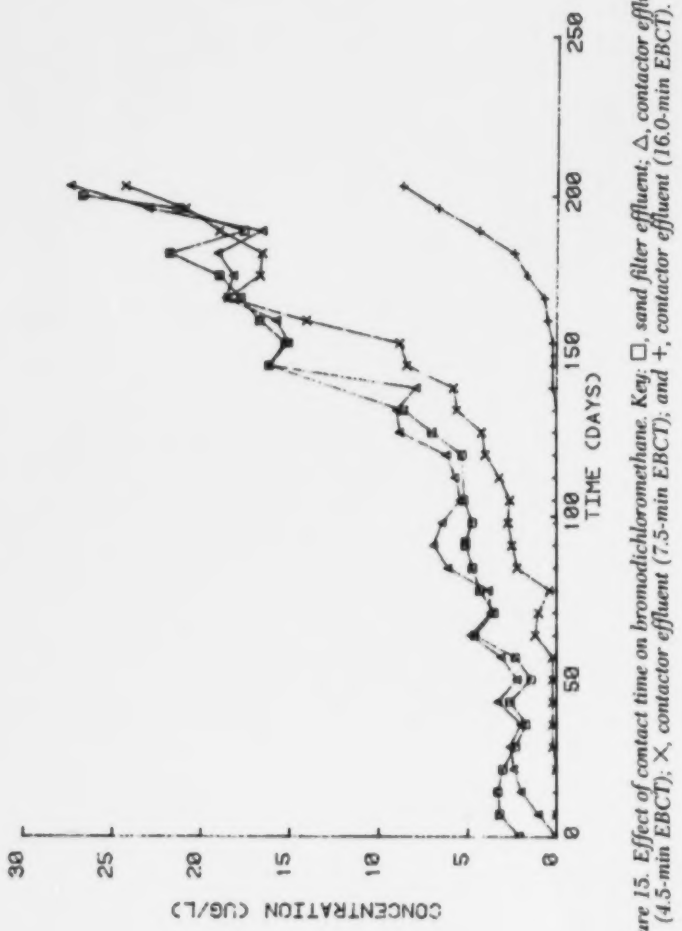

Figure 15. Effect of contact time on bromodichloromethane. Key: □, sand filter effluent; △, contactor effluent (4.5-min EBCT); ×, contactor effluent (7.5-min EBCT); and +, contactor effluent (16.0-min EBCT).

Table III. Carbon Analyses

Surface area (BET)
Iodine number
Modified phenol value
Molasses index
Volatile matter
Ash content
Apparent density
Abrasion number
Sieve analyses

performed on the carbon. Table IV shows the reduction in carbon capacity during the first carbon exposure for both the sand replacement and contactor carbon system using conventional laboratory carbon evaluation techniques. As expected, the surface area, iodine number, and decolorizing index decreased indicating a decrease in adsorptive capacity. An increase was shown, as expected, in the phenol value and volatiles, which also show decreased adsorptive capacity. The apparent density also shows an increase that might be attributed to adsorbed organics as well as inorganics. The increase in ash content indicates the inorganic residues, such as calcium, that were picked up by carbon during the exposure cycle. The ash results show the effect of a combined filtration/adsorption process as compared to a straight adsorption process. The percent of ash on the spent carbon from the sand replacement system was 2.7 times the value for the virgin carbon as compared to the factor of 1.2 shown for the contactor system. Although the ash values of the systems were restored to about virgin values for two cycles, as shown in Table IV, ash build-up will be watched in subsequent cycles.

Table IV. Results of Carbon Analyses

	Sand Replacement			*Contactor*		
	Virgin	*Spent*	*S/V*	*Virgin*	*Spent*	*S/V*
BET surface area (m^2/g)	1102	1052	0.95	1070	1030	0.96
Iodine number (mg/g)	1184	711	0.60	1128	824	0.73
Modified phenol value (ppm)	16.2	38.3	2.4	18	30.2	1.7
Molasses decolorizing index (DI units)	10.5	5.9	0.56	9.6	6.8	0.71
Ash (%)	7.06	19.1	2.7	8.1	9.46	1.2
Volatile matter (%)	2.49	11.0	4.4	3.57	9.82	2.8
Apparent density (g/mL)	0.483	0.600	1.2	0.506	0.558	1.1
Mean particle diameter	1.25	1.34	1.07	1.10	1.11	1.0

Carbon Reactivation. The 7.5-min EBCT systems generally exceed the previously cited reactivation cycle criteria in about 6–12 weeks depending on influent conditions, and the 0.22-m^3 (5-MGD) sand replacement systems were generally reactivated with that frequency. Each 0.04-m^3/s (1-MGD) contactor has a total EBCT of 16 min; thus, the time between reactivation cycles was longer, generally 3–6 months, depending on influent conditions. The 7.5-min EBCT was constantly monitored for comparison with the sand replacement system. The amount of carbon available for repetitive reactivation consisted of 40,909 kg (90,000 lb) in each of two sand replacement systems and 19,090 kg (42,000 lb) in each of the four contactor systems for a total of about 158,182 kg (348,000 lb).

The initial shakedown and early operation of the fluid bed reactivation followed the pattern described by many investigators. The major problems contributing to down time center around the following areas:

1. Instrumentation required to operate the system automatically. Temperature-related controllers such as thermocouples, recorders, and oxygen analyzers; computer hardware and software; and gas flow controllers.
2. Removing sand from dryer plates and unplugging dryer plate holes because the sand separator does not prevent sand carryover to the furnace.
3. Bearings for blowers and dewatering screw; bent screw shaft.
4. Design of cyclone separator discharge flap valve.

These items, plus many additive minor problems, predictable and unpredictable, have combined to result in a total worst case estimate of 43% operating time (57% down time) during the time period from April 10, 1980 through January 31, 1981. The down time for this period includes all time lost by unavailability of the reactivation system for any reactivator-related reasons. Problems related to the contactors and scheduled down time are not included. Thus, if carbon was scheduled for reactivation and a problem with the contactors prevented reactivation from taking place, then the reactivation system was not credited with down time. The above time period did include problems during shakedown and long time delays that now can be avoided by a proper equipment inventory and experiences learned to date. A more realistic, but still very conservative, estimate of 58% operating time or 42% down time experience is provided during the time period between July 1, 1980 and December 31, 1980. This period was after the initial 3 months of trials. The system operated 24 h/day for a 5-day week from October through December and adjustment for a 7-day week, 24-h/day operation would result in an additional 3% operating time to 60% or a projected down time of 40%. Experiences with similar systems at Hattiesburg, Miss. and Manchester, N.H. seem to forecast continued improvements in operating time.

The primary objective of the reactivation process was to return the spent carbon to nearly its virgin properties with a minimum amount of damage to the basic pore structures. Although all laboratory control tests listed in Table III were performed, the primary carbon quality control tests during the reactivation cycles were apparent density and iodine number. Various operating controls such as temperature, carbon feed rate, and steam feed rate were controlled closely during the reactivation process. Table V provides an example of furnace operating conditions experienced during the carbon feed; use averaged about 482 lb/h based on a backwash and drained density of 0.44 g/cc (27.5 lb/ft^3). Based on the apparent density of 0.48 g/cc (30 lb/ft^3) a carbon feed rate of about 243 kg/h (535 lb/h) was averaged to date. As expected, the duration of run for the 41-metric ton (45-ton) sand replacement carbon system was about twice the 19-metric ton (21-ton) contactor carbon system. The fuel requirements for the operating conditions shown are provided as an example and do not represent an average condition. However, they serve to illustrate the amount of fuel required by the incineration or afterburner function as compared to the total requirements. Fuel requirements were reported by other investigators for a multiple hearth and rotary kiln system used for reactivation of carbon used in wastewater applications at Pomona, Calif. (7, 8). The wastewater applications resulted in heavier organic loadings, but a general comparison of fuel requirement shows that about 19,000 kJ/kg (8000 BTU/lb) was used for the multiple hearth system and about 28,000 kJ/kg (12,000 BUT/lb) for the rotary kiln system as compared to about 14,000 kJ/kg (6000 BTU/lb) shown in Table V. The steam used per pound of carbon and typical operating temperatures are listed.

Table VI shows a comparison of carbon characteristics for the sand replacement carbon through the first two cycles of operation. Sufficient data have not been analyzed yet to draw firm conclusions regarding the effects of multiple reactivations on carbon adsorption capacity. However, the results shown would generally indicate that the carbon was more heavily loaded during the first cycle of operation and that the reactivation process is returning the spent carbon to very near the conditions present in the carbon at the start of each cycle. The analysis of regenerated carbon both before and after addition of make-up carbon (R-1B and R-1A) aids in assessment of the results achieved by each successive adsorption and reactivation cycle. Explaining the changes in iodine number or volatile matter, for example, is easier when the R-1A sample data are included. Without the R-1A iodine number of 1128 mg/g, the spent value in the

Table V. Example of Reactivation System Operating Conditions

Parameter	*Value*
Duration of run (h)	150–230 sand replacement
	90–110 contactor
Furnace loading	
kg carbon/h	219
lb carbon/h	482
Fuel used (BTU/lb carbon)	5973
incineration	3060
reactivation	1475
steam	1438
Steam used (lb/lb carbon)	0.6
Temperatures (°F)	
reactivation zone	1500
incineration zone	1600
dryer	300
dryer loop	250
scrubber stack	varies with water temperature
Total carbon losses	15.9% contactor
	18.9% sand replacement

Table VI. Effect of Reactivation on Sand Replacement Carbon

	First Cycle			*Second Cycle*		
Carbon Characteristics	*V*	*S*	*R-1B*	*R1-A*	*S*	*R-2B*
Surface area (m^2/g)	1102	1051	1022	983	1130	1119
Iodine number (mg/g)	1184	711	1011	1128	1093	1111
Modified phenol (ppm)	16.2	38.3	21.7	19.8	32.4	21.1
AWWA CALC (mg/L)	1.84	4.36	2.47	2.25	3.69	2.41
Molasses index (DI units)	10.5	5.9	10.8	13.5	11.2	14.5
Volatile matter (%)	2.49	11.0	3.87	4.79	10.66	5.38
Ash (%)	7.06	19.1	8.31	8.33	10.48	8.44
Apparent density (g/mL)	0.483	0.600	0.482	0.474	0.503	0.471
Mean particle diameter (mm)	1.251	1.336	1.178	1.165	0.932	0.945
Abrasion number (% by mean PD)	67.3	73.6	69.5	73.8	68.4	69.8

second cycle (1093) would appear to be higher than the R-1B value of 1011 from the reactivation in the first cycle. The first adsorption cycle lasted about 6 months, whereas the second cycle lasted only 6 weeks. Therefore, the large decrease from virgin to spent in the first cycle agrees with operating time conditions and likewise the small decrease from 1128 to 1093 mg/g during the second cycle is consistent with operating experience. The same effect is shown by the volatile matter, modified phenol value, and molasses decolorizing data. Also, the effect that virgin make-up carbon has on restoring some of the carbon characteristics may be estimated by comparing the values R-1B and R-1A.

Pore size distribution data will be compared with the general parameters such as iodine number and molasses decolorizing index (DI) to evaluate the relationship of these carbon control techniques to the changes in pore characteristics and performance for removing various specific organics. The increase in molasses DI, for instance, would indicate that there was a slight increase in the surface area in the pores greater than 30 Å. Pore size distribution data for the first cycle of data also showed a somewhat greater surface area between 35 Å and 150 Å. This increase may agree with the molasses DI data. Also, reduced iodine numbers and modified phenol values for cycle 1 could indicate a slight reduction in the surface area available in fine pores. If such a trend were to continue for more than the usual first few cycles, then eventually there would be an increased surface area available to large color body molecules. As mentioned previously, more data are required before a proper evaluation of differences can be performed.

Although the ash content increased by 170% during the first cycle of exposure to drinking water and 26% during the second cycle, the reactivation process returned the values to near virgin quality. The apparent density was the primary practical test used to control the carbon quality during the reactivation process. Frequent determinations were performed during each cycle and results in Table VI are typical of the close approximation of virgin conditions achieved. These results are especially good when the light loadings are taken into consideration. Such light organic loadings would be typical for carbon used at water supply utilities (*9–11*). There appears to be an overall reduction of mean particle diameter from the virgin value to the value shown for the second cycle regenerated carbon (R-2B). The abrasion numbers indicate no softening of particles has taken place thus far.

The measurement of carbon losses is an important consideration in the economic evaluation of the total carbon system installed. In this chapter the total carbon loss is defined as the difference in the spent carbon volume transferred to the spent tank and the regenerated carbon volume returned back to the system. All measurements were performed after backwashing. The sand replacement system volume measurements

include corrections for any volume of support sand removed and found by the sand separator system. The contactor system did not have a sand support and, thus, did not require such a correction. The total carbon losses are attributed to both the attrition losses caused by the transport system and carbon loss caused by the oxidation of fines in the transport process and oxidation of the surfaces of the carbon particles by the activating gases. Total carbon losses are shown in Tables V and VII. The range of losses can be seen from the data shown in Table VII. The losses ranged from about 14 to 19% for the sand replacement system. These comparisons, although not without a degree of error, provide an estimate of the total carbon losses experienced in the two systems compared. Such differences will effect the economic comparison that ultimately will be performed. The questions of where the losses occur and how to reduce them become important when trying to minimize operating costs. Attempts were made to measure the carbon losses between the spent and regenerated storage tanks. The confidence in these measurements is not as great as total system loss measurements, and only gross comparisons are planned. Table VIII shows the results achieved to date. Although we may be skeptical about some of the high and low values for each system, we do not have sufficient cause to say that one should disregard them. If a gross comparison is performed for the averages shown, it would appear that about 1% of the carbon is lost outside of the battery limits included by the spent tank to regenerated tank measurements. A special test was conducted whereby the carbon transport loss was measured by moving spent carbon from the contactor to the spent tank and then, with the rotary valve removed, the carbon was educted to

Table VII. Total Carbon Losses

Contactor			*Sand Replacement*		
ID	*Phase*	*Percent Loss*	*ID*	*Phase*	*Percent Loss*
C	3-0	15.8	21A	1-0	17.1
D	3-0	14.9	15A	3-0	22.6
A	3-0	13.8	21A	3-1	22.6
BB	3-0	18.0	15A	3-1	23.7
C	3-1	15.6	21A	3-2	13.7
D	3-1	18.9	15A	3-2	13.6
A	3-1	14.2			
Average loss		15.9	Average loss		18.9

Note: Total losses calculated by volume measurements in each carbon adsorption bed.

Table VIII. Carbon Losses Across Furnace

Contactor			*Sand Replacement*		
ID	*Phase*	*Percent Loss*	*ID*	*Phase*	*Percent Loss*
D	3-0	15.9	21A	3-1	31.2
A	3-0	12.0	15A	3-1	16.4
BB	3-0	14.6	21A	3-2	16.6
C	3-1	16.7	15A	3-2	8.7
D	3-1	3.4			
A	3-1	22.8			
Average loss		14.2	Average loss		18.2

Note: Furnace losses calculated by volume measurements in spent and regenerated carbon tanks.

the regeneration storage tank and then back to the contactor. The transport volume losses were 4%. A second loss measurement was conducted under normal conditions with the carbon passing through all of the usual stages which include blow case transport from the contactor vessel to the spent tank, rotary star valve feed to the screw conveyor, fluid bed transport through the normal furnace operating conditions, quench tank immersion, hydraulic eduction to the regenerated tank, and then back to the contactor. The total volume loss experienced was 15.6%; if the 4% transport loss is used as an estimate, then there was about an 11.6% volume loss between the rotary valve at the base of the spent tank and the eductor to the regenerated carbon storage tank.

This special test lends credence to the relative magnitude of the gross estimate of about 1% transport losses outside of the reactivation system battery limits. Also, the total loss for the special test was similar to the average shown for the contactors in Table VII. The estimate of an 11.6% "furnace" loss depends on the validity of the assumed 4% transport loss outside of the reactivation system limits as just described for the purpose of the loss measurements. Although many citations of losses appear, the actual reported data and specifics of measurements are sparse. Directo and coworkers (*7, 8*) reported total losses of about 7% for their experiences with the multiple hearth and rotary kiln. Nur (*12*) reported about 10% total losses and 2% transport losses for experiences with the infrared furnace at Pomona, Calif. Inhoffer and McGinnis (*13*) reported 5% losses for the infrared system at a water supply utility at Passaic, N.J. without any details of how the measurements were performed. Lykins (*14*) reported total losses most recently of about 14.8% (11% across the reactivation system) at the Manchester, N.H. water supply utility that uses a fluid bed reactivation system. The summation of carbon loss experience

described by Culp et al. (*15*) is generally without specifics of the measurement techniques used or specific data collected. However, the obvious conclusion one should arrive at is that general estimates of losses pose potential for misleading economic evaluations. Thus, the variation in impact of higher than estimated losses should be assessed and decisions made on the range of results presented.

Air Pollution Control Tests. During the normal reactivation cycle, atmospheric emission tests were performed on the fluid bed reactivation furnace. Triplicate tests were conducted downstream from the scrubber control device to determine the concentration and mass emission rates of particulate matter, nitrogen oxides (NO_x), and total gaseous nonmethane hydrocarbons. Also, tests were conducted downstream from the dryer cyclone to quantify particulate and volatile organic concentrations. In addition, the volumetric gas flow rate, temperature, moisture content, and composition (oxygen, carbon dioxide, and carbon monoxide) were measured. Where applicable, sample and analytical procedures used to determine particulates followed those described in EPA Method 5 of the *Federal Register* (*16*). The duct diameter at the sampling location did not meet the minimum required diameter of 30 cm (12 in.) specified in Method 1 of the *Federal Register* (*16*). Particulate samples were collected isokinetically at a point of average velocity in the duct. Velocity and temperature data were used to set isokinetic sample rates. The method was approved by the Regional Air Pollution Control Agency prior to testing. Table IX lists the average particulate emission data for the dryer loop and scrubber stack. Concentrations are reported in grains per dry standard cubic foot (gr/dscf) and emission rates in pounds per hour (lb/h). The filterable particulates represent the material collected in the sample probe and on the filter, both of which were heated to about 250°F. The condensible fraction represents that material which passed through the heated filter and was trapped in the impinger section of the sample train at a temperature of 68°F or less. The condensibles are not included as particulates for regulation in Ohio. The results are well within acceptable limits.

Table X lists the results of the organic emissions data collected during the testing. The high moisture content in the dryer loop test required the use of three filtered bag samples instead of the EPA Method 25 sampling tests that were used for the scrubber stack. Analytical procedures were those described in EPA Method 25 of the *Federal Register* (*17*) for data from both locations.

The nitrogen oxide (NO_x) tests were conducted according to procedures described in EPA Method 7 of the *Federal Register* (*18*). NO_x concentrations at the scrubber outlet averaged 97 parts per million with a mass emission rate of 0.37 lb/h. Further reseach related to air emission

Table IX. Particulate Emission Data

Type	*Dryer Loop*	*Scrubber Stack*
Concentration (gr./dscf)		
filterables[a]	0.5	0.0024
condensibles	0.039	0.0090
organic	0.018	—
inorganic	0.021	—
Mass emission rate (lb./h)		
filterables[a]	1.0	0.01
condensibles	0.08	0.043
organic	0.04	—
inorganic	0.04	—

[a]Only the filterables are considered for particulate emissions in Ohio.

testing may be conducted. All results thus far are well within acceptable limits imposed by Federal, State, and local regulations.

Costs

The general adsorption aspects of carbon have been long known and employed at Cincinnati Water Works by use of powdered activated carbon (PAC). Throughout the years, PAC has been used whenever river conditions warranted the use of an adsorbent. Thus, the study of a continuous operating GAC system and a potentially more effective and constant adsorption treatment barrier was a logical undertaking by water utilities. Since adsorption with carbon was already practiced, the emphasis for data gathering focused on the economic evaluation of the use of GAC systems and the constancy of protection provided for various time periods. The degree of organics removal obviously was equally important and the most comprehensive methods available were used as described elsewhere (*3*).

Table X. Total Gaseous Nonmethane Hydrocarbons

	Dryer Loop	*Scrubber Stack*
Concentration (ppm)[a]	3.1	219
Emission rate (lb/h)	0.002[b]	0.29[c]

[a]ppm is parts per million by volume, reported as CH_4.
[b]Emission rate for dryer is in pounds as CH_4 per hour, using measured concentration and average volumetric flow.
[c]Emission rate for scrubber is in pounds as CH_4 per hour, using standard conditions 16 lb/mol, and 385.4 ft^3/lb mol.

Tables XI and XII show some pertinent background data used in the cost calculations discussed later. The useful life cycles shown on Table XI were estimated by CWW personnel based on normal plant practices and operating experiences gained to date. Costs also were calculated using a straight 20-year life as is frequently found in existing literature. Table XII also contains some data used in the cost calculations. The range of power and fuel rates shows the effect of rapid escalation in energy costs. The actual rates and costs incurred were used in arriving at the capital and operating costs reported in subsequent cost tables.

Table XIII contains the costs to design and construct the carbon adsorption and reactivation system already described. No legal, fiscal and administrative, land, or insurance costs were included. However, all other typical design and construction costs, as well as resident engineering costs, are included. The capital costs of preparing two existing sand filters were minimal because no major alterations were performed. The reason that the initial carbon costs were about the same for the sand replacement and contactor system is because of the relative EBCTs previously described. In comparing total costs, one should keep in mind that a typical sand replacement system is being compared to a post-filtration adsorption system that used an EBCT thought to be fairly optimum.

No capital costs for a building in Table XIII or electric costs in Table XIV are shown for the sand replacement system because no additional costs were incurred above what was normally required for a sand filter. Thus, the costs shown reflect only the additional costs required by the use of either GAC system. The sand replacement transport system was manually fed and contributes to carbon losses through excess handling. In a full plant conversion, capital expenditures for a more efficient automatic transport system would appear to be a wise investment.

Table XI. Background Capital Cost Related Data (Useful Life Estimates for Units)

Sand Replacement System	
Initial carbon inventory	20-year life
Contactor System	
Initial carbon inventory	20-year life
Contactors and instrumentation	8-year life (likely conservative)
Building	25-year life
Reactivation System	
Furnace, tanks, and controls	15 years
Building	25 years

Note: All capital costs are amortized at 8% over the useful life span indicated.

Table XII. Criteria and Unit Costs

Unit costs		
Power	¢/KWH	2.19–2.91
Natural gas	%/100 ft^3	31.032–33.809
Water	%/1000 gallons	42.2
Carbon	¢/pound	46–50
Labor	$/hour	
Contactor and furnace		8
General labor		6.50
Maintenance		7–9
Criteria		
Furnace throughput	tons/yr	
Total		1200
Sand replacement		816
Contactor		384
Adsorption system throughput	MGY	
Sand replacement		2950
Contactors		1400
Natural gas fuel value	BTU/ft^3	1000
Carbon reactivation fuel use	BTU/lb C	6000

The capital costs and annual operating costs of the reactivation system are shown in Tables XIII and XIV along with the costs of the adsorption systems. Also, the capital and operations costs of the reactivation system were prorated between the adsorption systems based on the ratio of the carbon throughput requirements established for each adsorption system to the total throughput. Table XIV data show that the annual operating labor and material costs for the sand replacement system exceed that of the contactor system by more than the 2.5 ratio of the nominal rated capacities of the two systems.

These costs are primarily related to carbon transport and make-up carbon activities. The backwash criterion of 20-psi head loss (developed during pilot column tests) was never exceeded. Therefore, the contactor system pumping costs primarily consist of influent pumping because backwashing was not required during the normal course of contactor operation. The carbon transport costs are primarily the motive water requirements.

Table XV shows a comparison of the annual amortized capital costs based on the useful life cycles shown in Table XI and a straight 20-year life. The majoy difference appears in the contactor system data because of the 8-year life estimated for the entire contactor system. The total direct

Table XIII. Preliminary Capital Cost Estimates

	Adsorption		
	Sand Replacement (10 MGD)	*Contactors (4 MGD)*	*Reactivation (500 lb/h)*
Design and construction	3,400	832,000	740,200
Building cost	—	219,380	219,380
Initial carbon	96,250	84,000	—
Design and construction costs	99,650	1,135,380	959,580
Prorated reactivation cost to carbon systems	652,514	307,065	—
Design and construction total costs for compared systems	752,164	1,442,745	—

Note: Land, legal, fiscal and administrative, and insurance costs were not included.

Table XIV. Preliminary Annual Operating and Maintenance Costs (Dollars)

	Sand Replacement (10 MGD)	*Contactor (4 MGD)*	*Reactivation (500 lb/h)*
Carbon transport	2,530	1,000	23,000
Electric			
Pumping	—	14,700	—
Building	—	3,230	3,230
Furnace	—	—	6,900
Natural gas	—	—	51,400
Operating labor	66,640	16,240	85,400
Operating material	141,400	54,700	•
Maintenance labor	—	—	68,600
Maintenance material	—	—	1,500
	210,570	89,870	240,030
Prorated reactivation costs to carbon systems	163,220	76,810	—
	373,790	166,680	—

Table XV. Comparison of Annual Amortized Capital Costs (Dollars)

	Based on Table XI	*Based on 20-Year Life*
Sand replacement	10,002	10,002
Contactor	169,890	113,961
Reactivation	105,204	96,315
Prorated Reactivation		
Sand replacement	81,340	75,497
Contactor	203,556	144,782

and indirect costs are shown in Table XVI in dollars and Table XVII shows the costs in cents per 1000 gallons. The estimated total operating costs are about 7.5¢/1000 gallons for adsorption only for the sand replacement system and 14.5–18.5¢/1000 gallons for the contactor adsorption system. The criteria of a 6-week and 3-month cycle must be kept in mind when considering the costs shown. Relaxed criteria would result in lower costs with reduced water quality being delivered. The costs including prorated on-site reactivation are about 15¢/1000 gallons for the sand replacement system and 22–26¢/1000 gallons for the contactor system. The carbon reactivation costs were about 14¢/1b based on an annual carbon throughput of about 1200 tons.

The data collected during this study provide an estimate of the actual costs involved in design, construction, and operation of a system that could be full scale at some utilities. The costs in cents per 1000 gallons demonstrated for the small pressure contactor system or sand replacement system should provide an upper limit of costs for using GAC because both economy of scale and variation in limiting criteria will likely result in a reduced total operating cost for systems with a full-scale plant capacity similar to CWW.

The rest of the project will be directed at estimating whether inflation and long-term effects of high operating and low capital costs as compared to low operating and high capital costs will cause any change in the apparent lower total cost observed for the sand replacement system.

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Table XVI. Total Annual Operating Cost Estimates (Dollars)

	Sand Replacement (10 MGD)	*Contactor (4 MGD)*	*Reactivation*	*Sand Replacement + Reactivation*	*Contactor + Reactivation*
Direct operating cost/yr	210,570	89,870	240,030	373,790	166,680
Indirect operating cost/yr					
Based on Table XI data	10,002	169,890	105,204	81,340	203,556
(Based on 20-year life)	(10,002)	(113,961)	(96,315)	(75,497)	(144,782)
Total cost	220,572	259,760	345,234	455,130	370,236
(Total cost: based on 20-year life)	(220,572)	(203,831)	(336,345)	(449,287)	(311,462)

Table XVII. Total Annual Operating Cost Estimates (¢/1000 Gallons)

	Sand Replacement (10 MGD)	*Contactors (4 MGD)*	*Sand Replacement + Reactivation*	*Contactor + Reactivation*
Direct operating costs	7.2	6.4	12.7	11.9
Indirect operating costs				
Based on Table XI useful life data	0.3	12.1	2.7	14.5
Based on 20-year life data		(8.1)	(2.5)	(10.3)
Total operating costs				
Based on Table XI useful life data	7.5	18.5	15.4	26.4
Based on 20-year life data		(14.5)	(15.2)	(22.2)

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